Design and Analysis of a Process to Convert Methanol into Dimethyl Ether

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DESIGN AND ANALYSIS OF A PROCESS TO CONVERT METHANOL INTO DIMETHYL ETHER

By

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ABSTRACT

A recent economic downturn led to the loss of a contract for a company that sells methanol. The company has two options to recover some of the profit lost from the contract termination. One option is to sell the excess methanol at the spot market price, and the second option is to implement a process to convert the excess methanol into dimethyl ether. This project investigated the implementation of a process to convert the methanol from the lost contract into dimethyl ether. The process was simulated in AVEVA Process Simulation to estimate the size of equipment needed. A toller provided a list of equipment available to rent, that was similar in size to the estimated equipment in the base case, for the process consisting of three reactors, three distillation columns, and eight heat exchangers. Twelve viable equipment combinations consisting of one reactor, one column, and five heat exchangers each were identified. The least costly equipment set was determined through optimization of the viable equipment combinations. The annual cost of the rental and utilities for the most profitable combination was determined to be $662,000. The profit from the dimethyl ether produced was calculated to be $6.4 million dollars when the rental and utility costs for the process were deducted. This option would earn the company an extra $1.2 million in profit compared to selling the methanol on the open market for $5.2 million. Dimethyl ether production was concluded to be the most profitable option for the company.

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Introduction

A chemical company that produces commercial grade methanol has lost a long term contact due to a significant economic downturn in the last year. This leaves the company in a difficult situation. Two options have been proposed. The first option would be to sell the methanol on the open market, but due to a glut of methanol in the local market, the spot selling price for methanol is likely to be low. A second solution has been suggested involving the conversion of the excess methanol into dimethyl ether (DME). The DME production process shall occur on a rented, skid-mounted unit provided by a toller, who is an outside contractor. This installment is not permanent, and if the decision is made to halt production of DME, the toller has agreed to safely remove the equipment. There is a shortage of DME in the local market which means that the spot selling price for DME is higher than that of methanol.

Option three has the potential for a higher profit, due to the larger market demand for DME than for methanol. Design and analysis of the process to convert methanol into DME was completed to investigate the incentive for implementing this process.

DME Production Process

Dimethyl ether is produced by the catalytic dehydration of methanol. Vaporized methanol is fed into a packed bed reactor (PBR) of catalyst where the reversible reaction takes place as shown in Equation 1.

$$CH_3OH \leftrightarrow (CH_3)_2O + H_2O \qquad (1)$$

The other important unit operation in this process is distillation. The distillation column separates the water and unreacted methanol from the DME product. For this process to be successful, the single pass conversion of methanol in the reactor must reach 81% or higher, with a recycle stream being utilized for the unreacted methanol. This methanol will be returned to the methanol feed and sent back through the process. The product purity of DME must be 99.5% by weight. Dimethyl ether (DME) is used extensively in the chemical industry as an aerosol propellant for products such as hair spray and bug spray. Furthermore, it can be used in specially designed compression ignition diesel engines and creates less emissions than standard automotive fuels.

Environmental and Process Safety Considerations

Methanol and DME have separate safety and environmental considerations. First, the National Fire Protection Association (NFPA) diamonds were considered for both chemicals to determine the safety considerations for our operators and process. The diamonds are shown in Figure 1 [1, 2].

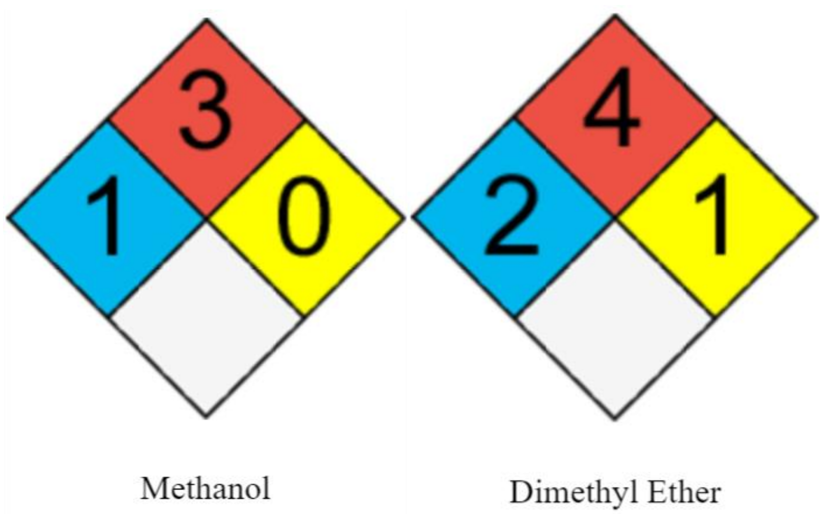

Figure 1: NFPA Diamonds for Methanol and DME

Methanol has a fire hazard of 3. This means that methanol has a flash point below 100°F. Methanol has a closed-cup flash point of 54°F and is highly flammable. Additionally, methanol has a health hazard of 1, which means that methanol slightly hazardous. Methanol is toxic if inhaled, ingested, or contacts skin. Additionally, methanol can cause adverse kidney and liver effects, can be fatal or cause blindness if swallowed, and can cause headache, nausea, and shortness of breath. Finally, methanol has a reactivity hazard of 0, meaning it is stable [3].

DME has a fire hazard of 4, which means it has a flash point below 73°F. DME is a gas at room temperature and can be stored in gas phase, or it can be stored as a liquid in a pressurized vessel much like propane. DME has a closed-cup flash point of -42°F and is highly flammable. Specific storage instructions should be discussed with a subject matter expert to ensure safe storage practices are in place. Additionally, DME has a health hazard of 2, which means it is hazardous. DME will cause nervous system depression if inhaled or swallowed and can cause drowsiness, dizziness, nausea or vomiting, headache, and unconsciousness. Finally, DME has a reactivity hazard of 1. This means DME can become unstable if heated. It should be noted that DME can react with oxygen to create explosive peroxides. When explosive peroxides are created or the DME is liquified, the DME becomes unstable when heated, but it is not unstable in pure form [4].

With these risks known, several safety precautions should be implemented to ensure safe handling of these chemicals. First, all operators and employees should wear proper personal protective equipment (PPE) when handling these chemicals or process equipment. This includes eye protection, gloves, and a respirator with an organic filter if proper ventilation is not available. Additionally, DME is peroxidizable, so DME should be stored in an opaque container, a peroxide inhibitor, such as hydroquinone, should be added to the DME storage tank, and a nitrogen blanket should be used in the storage tank to reduce peroxide formation. Furthermore, sources of ignition should be minimized, and proper fire safety should be implemented to reduce the risk of a fire and minimize the damage caused by a fire. Finally, methanol exists as a liquid at room temperature, and DME exists as a vapor at room temperature. Thus, storage requirements for these chemicals will differ, as methanol may be stored as a liquid in a vessel, but DME will either need to be compressed in gas cylinders or held in pressurized vessels for storage [3, 4].

DME has no known environmental hazards. However, it is recommended to consult local and federal guidelines for disposal. It is recommended that a flare system be utilized in the event DME must be vented to the environment. Methanol, however, is toxic to aquatic ecosystems and is highly mobile in soil and water. Thus, it is important that methanol not be released into the environment and is disposed of based on local and federal regulations [3, 4].

AVEVA Process Simulation

To further analyze and design the production of DME, the AVEVA Process Simulation software was utilized to simulate the process. This software assists calculations and economic evaluation necessary to determine the profitability/viability/feasibility of the proposed project through the consideration of operation conditions and process specifications. The creation of a simulation for the process allows for equipment sizing and operating cost estimations. The process of converting methanol to DME is demonstrated on a Process Flow Diagram (PFD) in the Appendix. The PFD displays the relationships between the major equipment within the plant.

The base case model was created within AVEVA and includes considerations for the thermodynamics of the system, reaction kinetics, and pressure-drop correlations. AVEVA enables the integration of equations directly into the simulation flowsheet without the need for a programming language. Thus, a given equation can be input in its original form, which makes it easily identifiable for the user. Submodels are programmed within the software to evaluate the economics of the process, displaying important values such as the annual operating costs or utility costs. AVEVA's Optimization Manager is a function that allows process optimization through an iterative calculation method. Certain variables can be changed and/or constrained in order to satisfy an objective function (i.e., minimize total utility costs or equivalent annual operating costs).

There are three simulation modes available in the program: Process, Fluid Flow, and Dynamics. Process mode is used to perform steady-state mass and energy balances that will allow for the design and sizing of process equipment. Fluid Flow mode can also be used for steady-state calculation but is primarily driven by the fluid dynamics within the system. It evaluates the sizes of valves, pipes, vessels, and heat exchangers. Dynamics mode evaluates processes over time and should be used to develop control systems that respond to deviations from the base case, such as a sudden change in flow, pressure, temperature, or liquid levels. For the creation of the base case model, the simulation was completed using the Process mode in order to focus on designing the key pieces of equipment for the DME process (reactor, column, and heat exchangers). This simplified the initial design to determine whether the project should move forward and be further investigated for profitability.

Base Case

The PFD of the process, shown in Figure A-1, was utilized to develop the base case simulation for the DME production facility. This base case simulation modeled unit operations for two major equipment groups. The reactor, a packed bed reactor (PBR), and the separations unit, a trayed distillation column, along with associated equipment, such as heat exchangers and pumps, to provide the necessary stream properties for the process streams.

The creation of the base case model was executed to accomodate the required annual feed rate of methanol (23,000 tonne/y), along with the recycled unreacted methanol from the process, the desired purity of DME (99.5 wt%), and the desired single-pass conversion of methanol to DME ($\geq$81%). This model served as a preliminary design that corresponds with the PFD.

To simulate the base case, each stream and unit operation shown on the PFD was placed in the flowsheet in order from left to right. First, the feed conditions were specified. The given temperature, pressure, and flowrate of methanol to the process were entered into the simulation. From there, the equipment was added and specified according to the given parameters and requirements. Submodels were used to calculate reaction kinetics in the reactor along with pressure drop correlations. Some important parameters for equipment were not specified, such as lengths and diameters, so they were calculated by the program based on the information that was added to the simulation. Once all the streams and equipment on the PFD were on the flowsheet, the simulation was solved with no error messages and all unspecified values were calculated by the program.

Next, the base case distillation column was optimized to complete the base case analysis. The initial design was optimized to minimize the Equivalent Annual Operating Costs (EAOC) associated with the distillation column, including the condenser and reboiler. Costs associated with the column's auxiliary equipment were considered negligible for these calculations. The EAOC was calculated as a function of the purchase costs of the distillation column, columns trays, the condenser, and the reboiler. The cost of the column is related to its volume, the cost of the trays is related to the column's cross-sectional area, and the cost of the exchangers (condenser and reboiler) are related to the required heat exchange area.

Parameters for feed tray location and number of column trays were changed to minimize the EAOC in order to determine the approximate minimum size of equipment that could be used for the process. These equipment specification and design estimates were compiled, as shown in Table A-1. The optimized operating point for the DME column is represented in Figure 2 as a 3D graph.

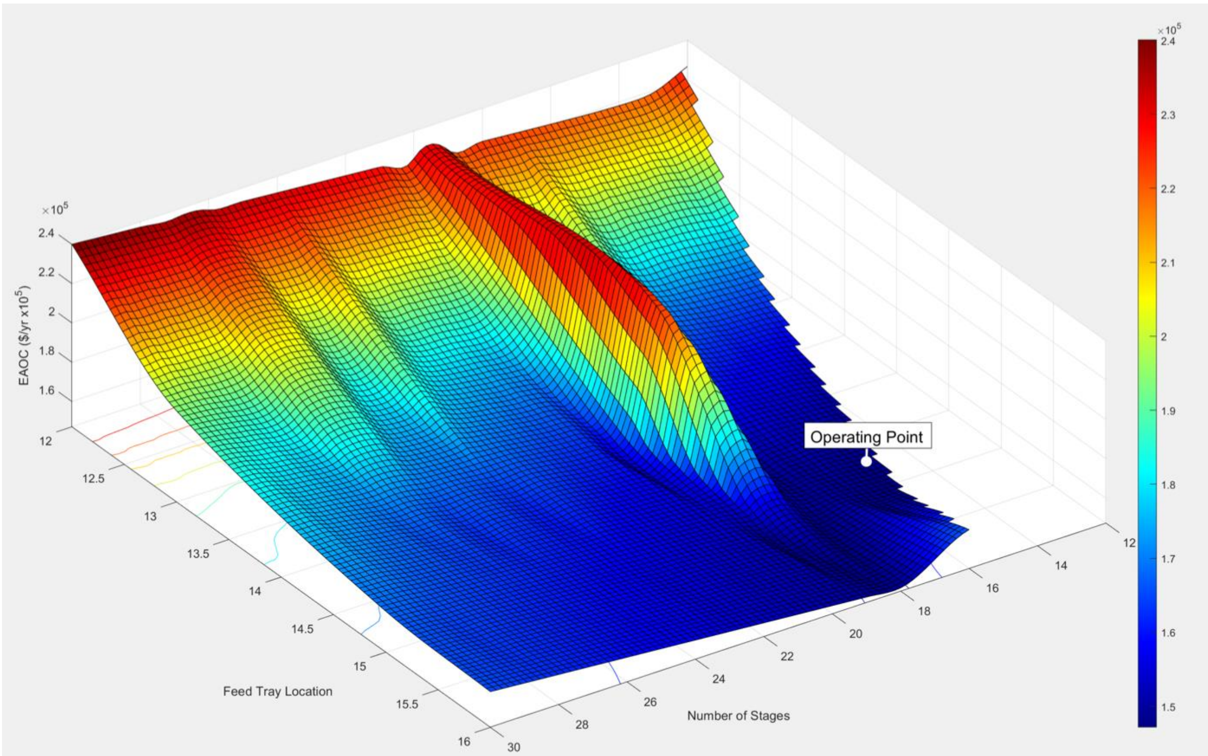

Figure 2: 3D Graph of DME Column Optimization Parameters: Number of Stages, Feed Location, and EAOC

The optimal column configuration had 15 trays with the feed coming into the column at tray 15. With the tower optimized, the base case was complete. Table A-2 shows the data for all process streams in the optimized base case.

Equipment Selection and Viability

After completion of the base case simulation and EAOC optimization, equipment sizing results were sent to the toller. From these results, the toller determined available equipment that could be utilized for the process. Three reactors, three columns, and eight heat exchangers were presented by the toller for consideration. The equipment presented is shown in Tables A-3a and A-3b. These include the size of the equipment and relevant equipment parameters, such as the maximum allowable working pressure (MAWP), maximum allowable working temperature (MAWT), and number of column trays. Additionally, some considerations were given by the toller to allow for simulation of these units while upholding the required operation parameters. These are shown in Table A-3a and are noted via asterisks to indicate the requirements and/or considerations.

To begin the simulation of the equipment, the possible reactor and column combinations available were considered first. As there were three reactors and three columns available, a maximum of nine reactor-column combinations could be created. These combinations were analyzed, without the specification of available heat exchangers, to determine if the combinations were viable with the required equipment specifications.

From this analysis, it was determined that column B would not be operable for the process. This was due to two interconnected reasons. The first was due to the MAWP of the column, which was 7 bar. As the process operates at 10 bar under base case operating conditions, it was found from the analysis that a MAWP of 7 bar for the distillation column was too low for operating procedures. Additionally, due to the low pressure in the distillation column, the condenser attached to the distillate portion of the column increased to a value that was unreasonably large for the process. With no heat exchanger large enough to accommodate for the size needed for the condenser, a secondary reason for the discontinuation of simulating with column B was provided.

With the elimination of column B, six combinations of reactors and columns were utilized in combination with the available heat exchangers to determine all viable combinations. By changing the heat exchangers, utilities used for the heat exchangers, and the stream conditions in the plant, twelve viable equipment combinations were identified.

Optimization of Equipment Combinations

Once the equipment combinations were modeled, each simulation was optimized to achieve the lowest possible utility costs. The cost of utilities included the electricity powering the two pumps as well as the cost of low, medium, and high-pressure steam, cooling water, and refrigerated cooling water used in the heat exchangers. Using the economic submodel in AVEVA Process Simulation, the hourly cost of the utilities was calculated. To perform the minimization of utility costs, the optimization function in AVEVA Process Simulation was used. In the optimization manager, the goal was set to minimize the cost of utilities. The variables for the optimization were process stream temperature into the column as well as pressure and tray diameter of the column, and the minimum and maximum allowable values for these variables was also specified. Additionally, constraints were set for the optimization based on the required flooding factor in the column being between 30% and 80% for each tray. The maximum allowable temperature within the column, as seen in Table A-3a, was also used as a constraint. The optimization was run for each of the 12 equipment combinations, and the minimum possible utility cost for each simulation was reached.

Economic Analysis

With the utility costs calculated for the 12 equipment combinations, the economic analysis was completed. The yearly rental costs of the equipment and the utility costs per year for each combination were added to determine the overall cost per year. Figures A-2a and A-2b show the breakdown of yearly costs for the equipment and utilities for each combination. The yearly cost for the 12 combinations ranged from $662,016 to $919,003 per year. Rental costs for equipment from the toller contributed more to the overall yearly cost than the utilities. The range in rental costs for the combinations was $278,280, while the range in utility costs was only $47,718. The yearly cost, from least to greatest for each of the 12 equipment combinations is shown in Table 1.

Table 1: Yearly Costs for Equipment Combinations

Equipment Combination	**Yearly Cost**	**Reactor/Column**	**Heat Exchangers***
1	$662,000	B/A	D,B,C,F,G
2	$678,000	A/A	D,B,C,A,G
3	$681,000	A/A	D,B,C,F,G
4	$803,000	C/A	F,B,C,D,G
5	$834,000	C/A	F,B,C,A,G
6	$835,000	C/A	A,B,C,H,G
7	$841,000	C/A	A,B,C,E,G
8	$857,000	C/C	A,B,C,D,G
9	$860,000	C/C	F,B,C,D,G
10	$889,000	A/C	F,B,C,A,G
11	$890,000	C/C	F,B,C,A,G
12	$919,000	C/C	A,B,C,E,G

*in order E-101, E-102, ..., E-105

Note that the exchangers chosen for E-102, E-103, and E-105 were the same for each combination. This is because, for E-102 and E-103, these exchangers were the only options that could operate at the high pressure and temperature, respectively. As for E-105, heat exchanger G was used because the smallest heat exchange area was needed for the reboiler.

The most expensive reactor and column combination was reactor C and column C at a price of $919,000 per year. This reactor and column combination, when paired with different heat exchanger options, made up four of the five most expensive combinations. Interestingly, the most expensive utility costs were found in the combination that was the least expensive overall. Reactor

B, column A was the equipment combination that was the least expensive at a price of $662,000 per year. This is largely due to the rental costs of reactor B and column A being the smallest out of the available equipment. Even though the utility costs for this configuration were the highest out of all the configurations, the use of the lowest cost reactor and column compensated for the larger utility cost. This equipment combination also achieved an 84% conversion of methanol, which exceeded the required minimum of 81% conversion.

The profit from selling the excess methanol on the open market compared to the profit if the excess methanol is converted into DME, after rental and utility costs are subtracted, is shown in Table 2:

Table 2: Potential Profit from Open Market Sale

Sale Type	**Profit ($M/yr)**
Open Market Methanol	5.2
Open Market DME	6.4

The company would increase their profit by $1.2 million per year if the DME process was implemented. This estimate, however, does not include deductions for costs like labor, maintenance, taxes, etc. More analysis would have to be done to determine the actual final profit when all costs are accounted for.

Conclusions and Recommendations

Converting the methanol from the lost contract into DME is a profitable option for the company. A base case model was successfully created to meet the required annual feed rate of methanol (23,000 tonne/y), the desired purity of DME (99.5 wt%), and the desired single-pass conversion of methanol to DME (≥81%). With this base case model, equipment sizing was estimated in order to obtain a list of equipment that could work for the process from the toller. Based on the equipment list provided by the toller, the DME conversion can be completed for an equipment rental and utility cost of $662,000 per year. A profit of $6.4 million per year can be made using the equipment combination shown in Table 2.

Before the process can be successfully implemented, there are two important steps that need to be taken. First, a dynamic simulation should be developed in AVEVA Process Simulation. In a dynamic simulation, a control system can be implemented and tested. The process needs to be tested for its ability to adapt to small changes in temperature, pressure, or flow rate that are common in production. If the system is unable to adapt quickly and adequately, the process is not viable. Second, a full net present value (NPV) analysis should be conducted to account for costs not covered in this project. This analysis will ensure that the process would still be profitable when every cost is considered.

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Appendix

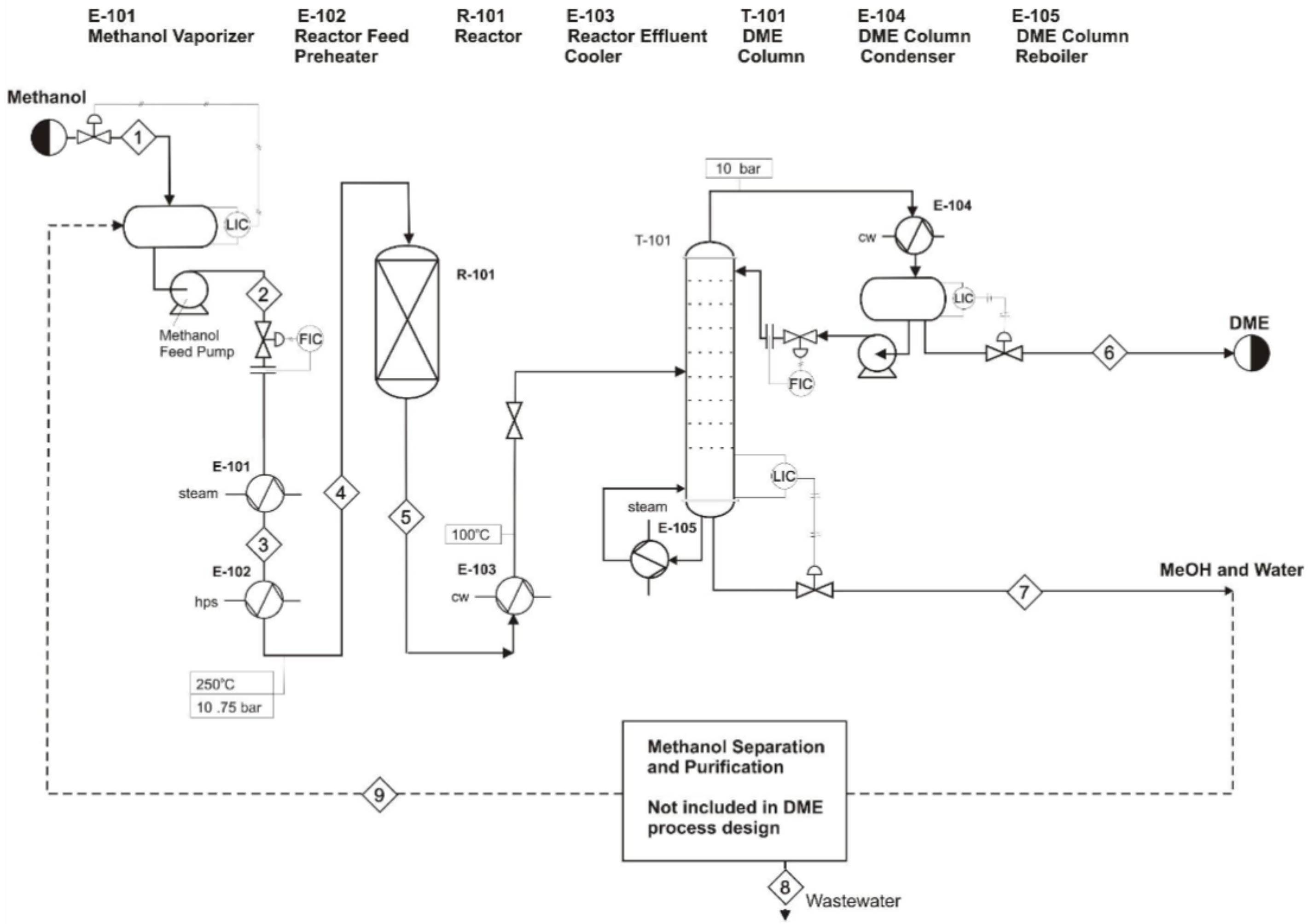

Figure A-1: Process Flow Diagram (PFD) for the conversion of methanol to dimethyl ether (DME)

Table A-1: Equipment Specifications for Optimized Base Case

Reactor (R-101)	L (m)	D (m)	Max Op Temp (°C)	Max Op Pressure (bar)	Catalyst Volume (m^3)		Pressure Drop (kPa)	
	7.09	1.00	370.7	10.8	3.34		15.1	
Column (T-101)	L (m)	D (m)	Max Op Temp (°C)	Max Op Pressure (bar)	#Trays	Feed Tray	Pressure Drop (kPa)	
	8.05	0.44	101.3	10.1	15	15	18.6	
Heat Exchangers	A (m^2)	Max-Tube		Max-Shell		Configuration S-Pass T-Pass	Process Side Pressure Drop (kPa)	Utility Side Pressure Drop (kPa)
		P (bar)	T (°C)	P(bar)	T(°C)			
Exchanger (E-101)	55.74	4.8	150.0	11.0	138.5	2-1	10	30
Exchanger (E-102)	8.68	46.9	260.0	10.9	250.0	2-1	10	30
Exchanger (E-103)	18.81	1.0	40.0	10.6	370.7	2-1	10	30
Exchanger (E-104)	24.68	1.0	15.0	10.0	46.2	2-1	5	30
Exchanger (E-105)	0.89	46.9	260.0	10.1	157.6	2-1	5	30

Table A-2: Stream Table for Optimized Base Case

	Stream 1	**Stream 2**	**Stream 3**	**Stream 4**	**Stream 5**	**Stream 6**	**Stream 7**	**Stream 8**	**Stream 9**
Temperature (°C)	30.00	30.19	138.53	250.00	370.69	30.00	146.43	65.06	30.00
Pressure (bar)	1.01	10.95	10.85	10.75	10.60	9.98	10.16	10.16	1.01
Vapor Fraction	0	0	1	1	1	0	0	0	0
Mass flow (tonne/hr)	2.66	3.28	3.28	3.28	3.28	1.91	1.37	0.75	0.62
Molar flow (kmol/hr)	75.43	92.92	92.92	92.92	92.92	37.78	55.13	37.65	17.49
Component flows									
DME (kmol/hr)	0.00	0.15	0.15	0.15	37.66	37.51	0.15	0.00	0.15
methanol (kmol/hr)	75.29	92.62	92.62	92.62	17.60	0.27	17.33	0.00	17.33
water (kmol/hr)	0.14	0.15	0.15	0.15	37.66	0.00	37.66	37.65	0.02

Table A-3a: Available Reactors and Columns

Equipment	Length (m)	Diameter (m)	Max Op Temp (°C)	Max Op Pressure (bar)	Catalyst Volume* (m^3)	Rental Cost ($/mo)**
Reactor A	5	1	400	12	3.93	10.0 k
Reactor B	4	0.8	400	12	2.00	6.31k
Reactor C	7	1.4	400	12	10.74	20.3 k
	Length*** (m)	Diameter**** (m)	Max Op Temp (°C)	Max Op Pressure (bar)	No. of Valve Trays*****	Rental Cost ($/mo.)
Column A	9	0.5	300	11	20	5.8 k
Column B	10	0.6	300	7	24	7.9 k
Column C	10	0.8	300	15	24	11.9 k

* This is maximum catalyst volume that can be accommodated in the reactor

** k indicates $1000

*** The column height reported by the toller accounts only for the total tray height. You can ignore the height of the sump in your design.

**** Flooding should not exceed 80%. If flooding is less than 30% then column can be retrofitted with small diameter trays at no extra cost to bring the flooding % up above 30%.

***** These are the number of actual trays – you my assume that the overall tray efficiency is 70%

Table A-3b: Available Heat Exchangers

	Area (m^2)	Max - Tube P(bar)/T(°C)	Max - Shell P(bar)/T(°C)	Configuration Shell-pass	Configuration Tube-pass	Rental Cost ($/mo.)
Exchanger A	125	15/150	15/150	1	2	5.9 k
Exchanger B	90	15/300	50/300	1	2	4.5 k
Exchanger C	60	15/150	15/400	1	2	3.7 k
Exchanger D	40	20/300	15/180	1	1	3.4 k
Exchanger E	180	20/300	50/300	1	2	6.7 k
Exchanger F	100	15/150	15/150	2	4	6.1 k
Exchanger G	20	20/300	15/180	1	1	1.1 k
Exchanger H	150	50/300	15/300	1	1	6.1 k

Equipment Set 1				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor B	$ 6,310	N/A	N/A
T-101	Column A	$ 5,800	N/A	N/A
E-101	Exchanger D	$ 3,400	LPS	$ 170,627
E-102	Exchanger B	$ 4,500	HPS	$ 39,205
E-103	Exchanger C	$ 3,700	CW	$ 14,839
E-104	Exchanger F	$ 6,100	Refrigerated CW	$ 25,941
E-105	Exchanger G	$ 1,100	MPS	$ 39,828
Pumps	N/A	N/A	Electricity	$ 656
	Yearly Costs	$ 370,920	-	$ 291,096
	Overall Yearly Cost			**$ 662,016**

Equipment Set 2				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor A	$ 10,000	N/A	N/A
T-101	Column A	$ 5,800	N/A	N/A
E-101	Exchanger D	$ 3,400	LPS	$ 171,513
E-102	Exchanger B	$ 4,500	HPS	$ 34,878
E-103	Exchanger C	$ 3,700	CW	$ 14,261
E-104	Exchanger A	$ 5,900	CW	$ 4,032
E-105	Exchanger G	$ 1,100	MPS	$ 39,560
Pumps	N/A	N/A	Electricity	$ 1,204
	Yearly Costs	$ 412,800	-	$ 265,449
	Overall Yearly Cost			**$ 678,249**

Equipment Set 3				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor A	$ 10,000	N/A	N/A
T-101	Column A	$ 5,800	N/A	N/A
E-101	Exchanger D	$ 3,400	LPS	$ 171,512
E-102	Exchanger B	$ 4,500	HPS	$ 34,878
E-103	Exchanger C	$ 3,700	CW	$ 14,246
E-104	Exchanger F	$ 6,100	CW	$ 4,011
E-105	Exchanger G	$ 1,100	MPS	$ 39,559
Pumps	N/A	N/A	Electricity	$ 1,212
	Yearly Costs	$ 415,200	-	$ 265,419
	Overall Yearly Cost			**$ 680,619**

Equipment Set 4				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column A	$ 5,800	N/A	N/A
E-101	Exchanger F	$ 6,100	LPS	$ 172,133
E-102	Exchanger B	$ 4,500	HPS	$ 35,019
E-103	Exchanger C	$ 3,700	CW	$ 12,382
E-104	Exchanger D	$ 3,400	CW	$ 2,659
E-105	Exchanger G	$ 1,100	MPS	$ 40,019
Pumps	N/A	N/A	Electricity	$ 1,671
	Yearly Costs	$ 538,800	-	$ 263,882
	Overall Yearly Cost			**$ 802,682**

Equipment Set 5				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column A	$ 5,800	N/A	N/A
E-101	Exchanger F	$ 6,100	LPS	$ 171,507
E-102	Exchanger B	$ 4,500	HPS	$ 34,914
E-103	Exchanger C	$ 3,700	CW	$ 14,274
E-104	Exchanger A	$ 5,900	CW	$ 4,018
E-105	Exchanger G	$ 1,100	MPS	$ 39,556
Pumps	N/A	N/A	Electricity	$ 1,201
	Yearly Costs	$ 568,800	-	$ 265,471
	Overall Yearly Cost			**$ 834,271**

Equipment Set 6				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column A	$ 5,800	N/A	N/A
E-101	Exchanger A	$ 5,900	LPS	$ 171,673
E-102	Exchanger B	$ 4,500	HPS	$ 34,287
E-103	Exchanger C	$ 3,700	CW	$ 14,810
E-104	Exchanger H	$ 6,100	CW	$ 4,621
E-105	Exchanger G	$ 1,100	MPS	$ 39,501
Pumps	N/A	N/A	Electricity	$ 1,389
	Yearly Costs	$ 568,800	-	$ 266,280
	Overall Yearly Cost			**$ 835,080**

Figure A-2a: Optimized Toller Equipment Combinations 1-6

Equipment Set 7				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column A	$ 5,800	N/A	N/A
E-101	Exchanger A	$ 5,900	LPS	$ 171,700
E-102	Exchanger B	$ 4,500	HPS	$ 34,291
E-103	Exchanger C	$ 3,700	CW	$ 14,281
E-104	Exchanger E	$ 6,700	CW	$ 4,022
E-105	Exchanger G	$ 1,100	MPS	$ 39,528
Pumps	N/A	N/A	Electricity	$ 1,270
	Yearly Costs	$ 576,000	-	$ 265,092
	Overall Yearly Cost			**$ 841,092**

Equipment Set 8				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column C	$ 11,900	N/A	N/A
E-101	Exchanger A	$ 5,900	LPS	$ 171,701
E-102	Exchanger B	$ 4,500	HPS	$ 34,256
E-103	Exchanger C	$ 3,700	CW	$ 12,298
E-104	Exchanger D	$ 3,400	CW	$ 4,224
E-105	Exchanger G	$ 1,100	MPS	$ 23,461
Pumps	N/A	N/A	Electricity	$ 1,437
	Yearly Costs	$ 609,600	-	$ 247,378
	Overall Yearly Cost			**$ 856,978**

Equipment Set 9				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column C	$ 11,900	N/A	N/A
E-101	Exchanger F	$ 6,100	LPS	$ 171,473
E-102	Exchanger B	$ 4,500	HPS	$ 34,992
E-103	Exchanger C	$ 3,700	CW	$ 12,328
E-104	Exchanger D	$ 3,400	CW	$ 4,223
E-105	Exchanger G	$ 1,100	MPS	$ 23,496
Pumps	N/A	N/A	Electricity	$ 1,344
	Yearly Costs	$ 612,000	-	$ 247,856
	Overall Yearly Cost			**$ 859,856**

Equipment Set 10				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor A	$ 10,000	N/A	N/A
T-101	Column C	$ 20,300	N/A	N/A
E-101	Exchanger F	$ 6,100	LPS	$ 171,768
E-102	Exchanger B	$ 4,500	HPS	$ 34,922
E-103	Exchanger C	$ 3,700	CW	$ 13,482
E-104	Exchanger A	$ 5,900	Refrigerated CW	$ 24,829
E-105	Exchanger G	$ 1,100	MPS	$ 23,747
Pumps	N/A	N/A	Electricity	$ 1,054
	Yearly Costs	$ 619,200	-	$ 269,803
	Overall Yearly Cost			**$ 889,003**

Equipment Set 11				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column C	$ 11,900	N/A	N/A
E-101	Exchanger F	$ 6,100	LPS	$ 171,487
E-102	Exchanger B	$ 4,500	HPS	$ 34,995
E-103	Exchanger C	$ 3,700	CW	$ 12,576
E-104	Exchanger A	$ 5,900	CW	$ 4,329
E-105	Exchanger G	$ 1,100	MPS	$ 23,510
Pumps	N/A	N/A	Electricity	$ 1,223
	Yearly Costs	$ 642,000	-	$ 248,120
	Overall Yearly Cost			**$ 890,120**

Equipment Set 12				
Equipment on PFD	**Toller's Equipment**	**Monthly Rental Cost**	**Utility Type**	**Yearly Utility Cost**
R-101	Reactor C	$ 20,300	N/A	N/A
T-101	Column C	$ 11,900	N/A	N/A
E-101	Exchanger A	$ 5,900	LPS	$ 171,768
E-102	Exchanger B	$ 4,500	HPS	$ 34,922
E-103	Exchanger C	$ 3,700	CW	$ 13,482
E-104	Exchanger E	$ 6,700	Refrigerated CW	$ 24,829
E-105	Exchanger G	$ 1,100	MPS	$ 23,747
Pumps	N/A	N/A	Electricity	$ 1,054
	Yearly Costs	$ 649,200	-	$ 269,803
	Overall Yearly Cost			**$ 919,003**

Figure A-2b: Optimized Toller Equipment Combinations 7-12